

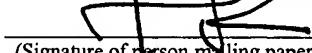
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TO ALL WHOM IT MAY CONCERN:

Be it known that WE, JOHN D. WILKINSON, JOE T. LYNCH, and HANK M. HUDSON, all citizens of the United States, all residing in Midland, County of Midland, State of Texas, whose post office addresses are 2800 W. Dengar, Midland, Texas 79705; 5510 Ashwood Ct., Midland, Texas 79707; and 2508 W. Sinclair, Midland, Texas 79705, respectively, and KYLE T. CUELLAR, a citizen of the United States, residing in Katy, County of Fort Bend, State of Texas, whose post office address is 1611 Cottage Point, Katy, Texas 77494, have invented an improvement in

NATURAL GAS LIQUEFACTION

of which the following is a

SPECIFICATION

BACKGROUND OF THE INVENTION

[0001] This invention relates to a process for processing natural gas or other methane-rich gas streams to produce a liquefied natural gas (LNG) stream that has a high methane purity and a liquid stream containing predominantly hydrocarbons heavier than methane.

[0002] Natural gas is typically recovered from wells drilled into underground reservoirs. It usually has a major proportion of methane, i.e., methane comprises at least 50 mole percent of the gas. Depending on the particular underground reservoir, the natural gas also contains relatively lesser amounts of heavier hydrocarbons such as ethane, propane, butanes, pentanes and the like, as well as water, hydrogen, nitrogen, carbon dioxide, and other gases.

[0003] Most natural gas is handled in gaseous form. The most common means for transporting natural gas from the wellhead to gas processing plants and thence to the natural gas consumers is in high pressure gas transmission pipelines. In a number of circumstances, however, it has been found necessary and/or desirable to liquefy the natural gas either for transport or for use. In remote locations, for instance, there is often no pipeline infrastructure that would allow for convenient transportation of the natural gas to market. In such cases, the much lower specific volume of LNG relative to natural gas in the gaseous state can greatly reduce transportation costs by allowing delivery of the LNG using cargo ships and transport trucks.

[0004] Another circumstance that favors the liquefaction of natural gas is for its use as a motor vehicle fuel. In large metropolitan areas, there are fleets of buses, taxi cabs, and trucks that could be powered by LNG if there were an economic source of LNG available. Such LNG-fueled vehicles produce considerably less air pollution due to the clean-burning nature of natural gas when compared to similar vehicles powered by gasoline and diesel engines which combust higher molecular weight hydrocarbons. In addition, if the LNG is of high purity (i.e., with a methane purity of 95 mole percent or higher), the amount of carbon dioxide (a

"greenhouse gas") produced is considerably less due to the lower carbon:hydrogen ratio for methane compared to all other hydrocarbon fuels.

[0005] The present invention is generally concerned with the liquefaction of natural gas while producing as a co-product a liquid stream consisting primarily of hydrocarbons heavier than methane, such as natural gas liquids (NGL) composed of ethane, propane, butanes, and heavier hydrocarbon components, liquefied petroleum gas (LPG) composed of propane, butanes, and heavier hydrocarbon components, or condensate composed of butanes and heavier hydrocarbon components. Producing the co-product liquid stream has two important benefits: the LNG produced has a high methane purity, and the co-product liquid is a valuable product that may be used for many other purposes. A typical analysis of a natural gas stream to be processed in accordance with this invention would be, in approximate mole percent, 84.2% methane, 7.9% ethane and other C₂ components, 4.9% propane and other C₃ components, 1.0% iso-butane, 1.1% normal butane, 0.8% pentanes plus, with the balance made up of nitrogen and carbon dioxide. Sulfur containing gases are also sometimes present.

[0006] There are a number of methods known for liquefying natural gas. For instance, see Finn, Adrian J., Grant L. Johnson, and Terry R. Tomlinson, "LNG Technology for Offshore and Mid-Scale Plants", Proceedings of the Seventy-Ninth Annual Convention of the Gas Processors Association, pp. 429-450, Atlanta, Georgia, March 13-15, 2000 and Kikkawa, Yoshitsugi, Masaaki Ohishi, and Noriyoshi Nozawa, "Optimize the Power System of Baseload LNG Plant", Proceedings of the Eightieth Annual Convention of the Gas Processors Association, San Antonio, Texas, March 12-14, 2001 for surveys of a number of such processes. U.S. Pat. Nos. 4,445,917; 4,525,185; 4,545,795; 4,755,200; 5,291,736; 5,363,655; 5,365,740; 5,600,969;

5,615,561; 5,651,269; 5,755,114; 5,893,274; 6,014,869; 6,053,007; 6,062,041; 6,119,479; 6,125,653; 6,250,105 B1; 6,269,655 B1; 6,272,882 B1; 6,308,531 B1; 6,324,867 B1; 6,347,532 B1; PCT Patent Application No. WO 01/88447; and our co-pending U.S. Patent Application Serial Nos. 10/161,780 filed June 4, 2002 and 10/278,610 filed October 23, 2002 also describe relevant processes. These methods generally include steps in which the natural gas is purified (by removing water and troublesome compounds such as carbon dioxide and sulfur compounds), cooled, condensed, and expanded. Cooling and condensation of the natural gas can be accomplished in many different manners. "Cascade refrigeration" employs heat exchange of the natural gas with several refrigerants having successively lower boiling points, such as propane, ethane, and methane. As an alternative, this heat exchange can be accomplished using a single refrigerant by evaporating the refrigerant at several different pressure levels. "Multi-component refrigeration" employs heat exchange of the natural gas with one or more refrigerant fluids composed of several refrigerant components in lieu of multiple single-component refrigerants. Expansion of the natural gas can be accomplished both isenthalpically (using Joule-Thomson expansion, for instance) and isentropically (using a work-expansion turbine, for instance).

[0007] Regardless of the method used to liquefy the natural gas stream, it is common to require removal of a significant fraction of the hydrocarbons heavier than methane before the methane-rich stream is liquefied. The reasons for this hydrocarbon removal step are numerous, including the need to control the heating value of the LNG stream, and the value of these heavier hydrocarbon components as products in their own right. Unfortunately, little attention has been focused heretofore on the efficiency of the hydrocarbon removal step.

[0008] In accordance with the present invention, it has been found that careful integration of the hydrocarbon removal step into the LNG liquefaction process can produce both LNG and a separate heavier hydrocarbon liquid product using significantly less energy than prior art processes. The present invention, although applicable at lower pressures, is particularly advantageous when processing feed gases in the range of 400 to 1500 psia [2,758 to 10,342 kPa(a)] or higher.

[0009] For a better understanding of the present invention, reference is made to the following examples and drawings. Referring to the drawings:

[0010] FIG. 1 is a flow diagram of a natural gas liquefaction plant adapted for co-production of NGL in accordance with the present invention;

[0011] FIG. 2 is a pressure-enthalpy phase diagram for methane used to illustrate the advantages of the present invention over prior art processes; and

[0012] FIGS. 3, 4, 5, 6, 7, and 8 are flow diagrams of alternative natural gas liquefaction plants adapted for co-production of a liquid stream in accordance with the present invention.

[0013] In the following explanation of the above figures, tables are provided summarizing flow rates calculated for representative process conditions. In the tables appearing herein, the values for flow rates (in moles per hour) have been rounded to the nearest whole number for convenience. The total stream rates shown in the tables include all non-hydrocarbon components and hence are generally larger than the sum of the stream flow rates for the hydrocarbon components. Temperatures indicated are approximate values rounded to the nearest degree. It should also be noted that the process design calculations performed for the purpose of comparing the processes depicted in the figures are based on the assumption of no heat leak from

(or to) the surroundings to (or from) the process. The quality of commercially available insulating materials makes this a very reasonable assumption and one that is typically made by those skilled in the art.

[0014] For convenience, process parameters are reported in both the traditional British units and in the units of the International System of Units (SI). The molar flow rates given in the tables may be interpreted as either pound moles per hour or kilogram moles per hour. The energy consumptions reported as horsepower (HP) and/or thousand British Thermal Units per hour (MBTU/Hr) correspond to the stated molar flow rates in pound moles per hour. The energy consumptions reported as kilowatts (kW) correspond to the stated molar flow rates in kilogram moles per hour. The production rates reported as pounds per hour (Lb/Hr) correspond to the stated molar flow rates in pound moles per hour. The production rates reported as kilograms per hour (kg/Hr) correspond to the stated molar flow rates in kilogram moles per hour.

DESCRIPTION OF THE INVENTION

[0015] Referring now to FIG. 1, we begin with an illustration of a process in accordance with the present invention where it is desired to produce an NGL co-product containing about one-half of the ethane and the majority of the propane and heavier components in the natural gas feed stream. In this simulation of the present invention, inlet gas enters the plant at 90°F [32°C] and 1285 psia [8,860 kPa(a)] as stream 31. If the inlet gas contains a concentration of carbon dioxide and/or sulfur compounds which would prevent the product streams from meeting specifications, these compounds are removed by appropriate pretreatment of the feed gas (not illustrated). In addition, the feed stream is usually dehydrated to prevent hydrate (ice) formation under cryogenic conditions. Solid desiccant has typically been used for this purpose.

[0016] The feed stream 31 is cooled in heat exchanger 10 by heat exchange with refrigerant streams and flashed separator liquids at -44°F [-42°C] (stream 39a). Note that in all cases heat exchanger 10 is representative of either a multitude of individual heat exchangers or a single multi-pass heat exchanger, or any combination thereof. (The decision as to whether to use more than one heat exchanger for the indicated cooling services will depend on a number of factors including, but not limited to, inlet gas flow rate, heat exchanger size, stream temperatures, etc.) The cooled stream 31a enters separator 11 at 0°F [-18°C] and 1278 psia [8,812 kPa(a)] where the vapor (stream 32) is separated from the condensed liquid (stream 33).

[0017] The vapor (stream 32) from separator 11 is divided into two streams, 34 and 36, with stream 34 containing about 15% of the total vapor. Some circumstances may favor combining stream 34 with some portion of the condensed liquid (stream 38) to form combined stream 35, but in this simulation there is no flow in stream 38. Stream 35 passes through heat exchanger 13 in heat exchange relation with refrigerant stream 71e and liquid distillation stream 40, resulting in cooling and substantial condensation of stream 35a. The substantially condensed stream 35a at -109°F [-78°C] is then flash expanded through an appropriate expansion device, such as expansion valve 14, to the operating pressure (approximately 465 psia [3,206 kPa(a)]) of fractionation tower 19. During expansion a portion of the stream is vaporized, resulting in cooling of the total stream. In the process illustrated in FIG. 1, the expanded stream 35b leaving expansion valve 14 reaches a temperature of -125°F [-87°C] and is then supplied at an upper mid-point feed position in absorbing section 19a of fractionation tower 19.

[0018] The remaining 85% of the vapor from separator 11 (stream 36) enters a work expansion machine 15 in which mechanical energy is extracted from this portion of the high

pressure feed. The machine **15** expands the vapor substantially isentropically to the tower operating pressure, with the work expansion cooling the expanded stream **36a** to a temperature of approximately -76°F [-60°C]. The typical commercially available expanders are capable of recovering on the order of 80-85% of the work theoretically available in an ideal isentropic expansion. The work recovered is often used to drive a centrifugal compressor (such as item **16**) that can be used to re-compress the tower overhead gas (stream **49**), for example. The expanded and partially condensed stream **36a** is supplied as feed to absorbing section **19a** in distillation column **19** at a lower mid-column feed point. Stream **39**, the remaining portion of the separator liquid (stream **33**) is flash expanded to slightly above the operating pressure of demethanizer **19** by expansion valve **12**, cooling stream **39** to -44°F [-42°C] (stream **39a**) before it provides cooling to the incoming feed gas as described earlier. Stream **39b**, now at 85°F [29°C], then enters stripping section **19b** in demethanizer **19** at a second lower mid-column feed point.

[0019] The demethanizer in fractionation tower **19** is a conventional distillation column containing a plurality of vertically spaced trays, one or more packed beds, or some combination of trays and packing. As is often the case in natural gas processing plants, the fractionation tower may consist of two sections. The upper absorbing (rectification) section **19a** contains the trays and/or packing to provide the necessary contact between the vapor portion of the expanded stream **36a** rising upward and cold liquid falling downward to condense and absorb the ethane, propane, and heavier components; and the lower, stripping section **19b** contains the trays and/or packing to provide the necessary contact between the liquids falling downward and the vapors rising upward. The stripping section also includes one or more reboilers (such as reboiler **20**) which heat and vaporize a portion of the liquids flowing down the column to provide the

stripping vapors which flow up the column to strip the liquid product, stream **41**, of methane and lighter components. The liquid product stream **41** exits the bottom of demethanizer **19** at 150°F [66°C], based on a typical specification of a methane to ethane ratio of 0.020:1 on a molar basis in the bottom product. The overhead distillation vapor stream **37**, containing predominantly methane and lighter components, leaves the top of demethanizer **19** at -108°F [-78°C].

[0020] A portion of the distillation vapor (stream **42**) is withdrawn from the upper region of stripping section **19b**. This stream is cooled from -58°F [-50°C] to -109°F [-78°C] and partially condensed (stream **42a**) in heat exchanger **13** by heat exchange with refrigerant stream **71e** and liquid distillation stream **40**. The operating pressure in reflux separator **22** (461 psia [3,182 kPa(a)]) is maintained slightly below the operating pressure of demethanizer **19**. This provides the driving force which causes distillation vapor stream **42** to flow through heat exchanger **13** and thence into the reflux separator **22** wherein the condensed liquid (stream **44**) is separated from any uncondensed vapor (stream **43**). Stream **43** combines with the distillation vapor stream (stream **37**) leaving the upper region of absorbing section **19a** of demethanizer **19** to form cold residue gas stream **47** at -108°F [-78°C].

[0021] The condensed liquid (stream **44**) is pumped to higher pressure by pump **23**, whereupon stream **44a** at -109°F [-78°C] is divided into two portions. One portion, stream **45**, is routed to the upper region of absorbing section **19a** of demethanizer **19** to serve as the cold liquid that contacts the vapors rising upward through the absorbing section. The other portion is supplied to the upper region of stripping section **19b** of demethanizer **19** as reflux stream **46**.

[0022] Liquid distillation stream **40** is withdrawn from a lower region of absorbing section **19a** of demethanizer **19** and is routed to heat exchanger **13** where it is heated as it

provides cooling of distillation vapor stream **42**, combined stream **35**, and refrigerant (stream **71a**). The liquid distillation stream is heated from -79°F [-62°C] to -20°F [-29°C], partially vaporizing stream **40a** before it is supplied as a mid-column feed to stripping section **19b** in demethanizer **19**.

[0023] The cold residue gas (stream **47**) is warmed to 94°F [34°C] in heat exchanger **24**, and a portion (stream **48**) is then withdrawn to serve as fuel gas for the plant. (The amount of fuel gas that must be withdrawn is largely determined by the fuel required for the engines and/or turbines driving the gas compressors in the plant, such as refrigerant compressors **64**, **66**, and **68** in this example.) The remainder of the warmed residue gas (stream **49**) is compressed by compressor **16** driven by expansion machines **15**, **61**, and **63**. After cooling to 100°F [38°C] in discharge cooler **25**, stream **49b** is further cooled to -93°F [-69°C] (stream **49c**) in heat exchanger **24** by cross exchange with cold residue gas stream **47**.

[0024] Stream **49c** then enters heat exchanger **60** and is further cooled by expanded refrigerant stream **71d** to -256°F [-160°C] to condense and subcool it, whereupon it enters a work expansion machine **61** in which mechanical energy is extracted from the stream. The machine **61** expands liquid stream **49d** substantially isentropically from a pressure of about 638 psia [4,399 kPa(a)] to the LNG storage pressure (15.5 psia [107 kPa(a)]), slightly above atmospheric pressure. The work expansion cools the expanded stream **49e** to a temperature of approximately -257°F [-160°C], whereupon it is then directed to the LNG storage tank **62** which holds the LNG product (stream **50**).

[0025] All of the cooling for stream **49c** and a portion of the cooling for streams **35** and **42** is provided by a closed cycle refrigeration loop. The working fluid for this refrigeration cycle

is a mixture of hydrocarbons and nitrogen, with the composition of the mixture adjusted as needed to provide the required refrigerant temperature while condensing at a reasonable pressure using the available cooling medium. In this case, condensing with cooling water has been assumed, so a refrigerant mixture composed of nitrogen, methane, ethane, propane, and heavier hydrocarbons is used in the simulation of the FIG. 1 process. The composition of the stream, in approximate mole percent, is 6.9% nitrogen, 40.8% methane, 37.8% ethane, and 8.2% propane, with the balance made up of heavier hydrocarbons.

[0026] The refrigerant stream 71 leaves discharge cooler 69 at 100°F [38°C] and 607 psia [4,185 kPa(a)]. It enters heat exchanger 10 and is cooled to -15°F [-26°C] and partially condensed by the partially warmed expanded refrigerant stream 71f and by other refrigerant streams. For the FIG. 1 simulation, it has been assumed that these other refrigerant streams are commercial-quality propane refrigerant at three different temperature and pressure levels. The partially condensed refrigerant stream 71a then enters heat exchanger 13 for further cooling to -109°F [-78°C] by partially warmed expanded refrigerant stream 71e, further condensing the refrigerant (stream 71b). The refrigerant is condensed and then subcooled to -256°F [-160°C] in heat exchanger 60 by expanded refrigerant stream 71d. The subcooled liquid stream 71c enters a work expansion machine 63 in which mechanical energy is extracted from the stream as it is expanded substantially isentropically from a pressure of about 586 psia [4,040 kPa(a)] to about 34 psia [234 kPa(a)]. During expansion a portion of the stream is vaporized, resulting in cooling of the total stream to -262°F [-163°C] (stream 71d). The expanded stream 71d then reenters heat exchangers 60, 13, and 10 where it provides cooling to stream 49c, stream 35, stream 42, and the refrigerant (streams 71, 71a, and 71b) as it is vaporized and superheated.

[0027] The superheated refrigerant vapor (stream 71g) leaves heat exchanger 10 at 93°F [34°C] and is compressed in three stages to 617 psia [4,254 kPa(a)]. Each of the three compression stages (refrigerant compressors 64, 66, and 68) is driven by a supplemental power source and is followed by a cooler (discharge coolers 65, 67, and 69) to remove the heat of compression. The compressed stream 71 from discharge cooler 69 returns to heat exchanger 10 to complete the cycle.

[0028] A summary of stream flow rates and energy consumption for the process illustrated in FIG. 1 is set forth in the following table:

Table I

(FIG. 1)

Stream Flow Summary - Lb. Moles/Hr [kg moles/Hr]

<u>Stream</u>	<u>Methane</u>	<u>Ethane</u>	<u>Propane</u>	<u>Butanes+</u>	<u>Total</u>
31	40,977	3,861	2,408	1,404	48,656
32	38,538	3,336	1,847	830	44,556
33	2,439	525	561	574	4,100
34	5,781	501	277	125	6,683
36	32,757	2,835	1,570	705	37,873
40	3,896	2,170	1,847	829	8,742
42	8,045	1,850	26	0	9,922
43	4,551	240	1	0	4,792
44	3,494	1,610	25	0	5,130
45	1,747	805	12	0	2,565
46	1,747	805	13	0	2,565
37	36,393	1,970	11	0	38,380
41	33	1,651	2,396	1,404	5,484
47	40,944	2,210	12	0	43,172
48	2,537	137	1	0	2,676
50	38,407	2,073	11	0	40,496

Recoveries in NGL*

Ethane	42.75%
Propane	99.53%
Butanes+	100.00%
Production Rate	246,263 Lb/Hr [246,263 kg/Hr]

LNG Product

Production Rate	679,113 Lb/Hr [679,113 kg/Hr]
Purity*	94.84%
Lower Heating Value	946.0 BTU/SCF [35.25 MJ/m ³]

Power

Refrigerant Compression	94,868 HP [155,962 kW]
Propane Compression	25,201 HP [41,430 kW]
Total Compression	120,069 HP [197,392 kW]

Utility Heat

Demethanizer Reboiler	24,597 MBTU/Hr [15,888 kW]
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* (Based on un-rounded flow rates)

[0029] The efficiency of LNG production processes is typically compared using the "specific power consumption" required, which is the ratio of the total refrigeration compression power to the total liquid production rate. Published information on the specific power consumption for prior art processes for producing LNG indicates a range of 0.168 HP-Hr/Lb [0.276 kW-Hr/kg] to 0.182 HP-Hr/Lb [0.300 kW-Hr/kg], which is believed to be based on an on-stream factor of 340 days per year for the LNG production plant. On this same basis, the

specific power consumption for the FIG. 1 embodiment of the present invention is 0.139 HP-Hr/Lb [0.229 kW-Hr/kg], which gives an efficiency improvement of 21-31% over the prior art processes.

[0030] There are two primary factors that account for the improved efficiency of the present invention. The first factor can be understood by examining the thermodynamics of the liquefaction process when applied to a high pressure gas stream such as that considered in this example. Since the primary constituent of this stream is methane, the thermodynamic properties of methane can be used for the purposes of comparing the liquefaction cycle employed in the prior art processes versus the cycle used in the present invention. FIG. 2 contains a pressure-enthalpy phase diagram for methane. In most of the prior art liquefaction cycles, all cooling of the gas stream is accomplished while the stream is at high pressure (path A-B), whereupon the stream is then expanded (path B-C) to the pressure of the LNG storage vessel (slightly above atmospheric pressure). This expansion step may employ a work expansion machine, which is typically capable of recovering on the order of 75-80% of the work theoretically available in an ideal isentropic expansion. In the interest of simplicity, fully isentropic expansion is displayed in FIG. 2 for path B-C. Even so, the enthalpy reduction provided by this work expansion is quite small, because the lines of constant entropy are nearly vertical in the liquid region of the phase diagram.

[0031] Contrast this now with the liquefaction cycle of the present invention. After partial cooling at high pressure (path A-A'), the gas stream is work expanded (path A'-A'') to an intermediate pressure. (Again, fully isentropic expansion is displayed in the interest of simplicity.) The remainder of the cooling is accomplished at the intermediate pressure (path

A"-B'), and the stream is then expanded (path B'-C) to the pressure of the LNG storage vessel. Since the lines of constant entropy slope less steeply in the vapor region of the phase diagram, a significantly larger enthalpy reduction is provided by the first work expansion step (path A'-A") of the present invention. Thus, the total amount of cooling required for the present invention (the sum of paths A-A' and A"-B') is less than the cooling required for the prior art processes (path A-B), reducing the refrigeration (and hence the refrigeration compression) required to liquefy the gas stream.

[0032] The second factor accounting for the improved efficiency of the present invention is the superior performance of hydrocarbon distillation systems at lower operating pressures. The hydrocarbon removal step in most of the prior art processes is performed at high pressure, typically using a scrub column that employs a cold hydrocarbon liquid as the absorbent stream to remove the heavier hydrocarbons from the incoming gas stream. Operating the scrub column at high pressure is not very efficient, as it results in the co-absorption of a significant fraction of the methane from the gas stream, which must subsequently be stripped from the absorbent liquid and cooled to become part of the LNG product. In the present invention, the hydrocarbon removal step is conducted at the intermediate pressure where the vapor-liquid equilibrium is much more favorable, resulting in very efficient recovery of the desired heavier hydrocarbons in the co-product liquid stream.

Other Embodiments

[0033] One skilled in the art will recognize that the present invention can be adapted for use with all types of LNG liquefaction plants to allow co-production of an NGL stream, an LPG stream, or a condensate stream, as best suits the needs at a given plant location. Further, it will

be recognized that a variety of process configurations may be employed for recovering the liquid co-product stream. The present invention can be adapted to recover an NGL stream containing a significantly higher fraction of the C₂ components present in the feed gas, to recover an LPG stream containing only the C₃ and heavier components present in the feed gas, or to recover a condensate stream containing only the C₄ and heavier components present in the feed gas, rather than producing an NGL co-product containing only a moderate fraction of the C₂ components as described earlier. The present invention is particularly advantageous over the prior art processes when only partial recovery of the C₂ components in the feed gas is desired while capturing essentially all of the C₃ and heavier components, as the reflux stream **45** in the FIG. 1 embodiment allows maintaining very high C₃ component recovery regardless of the C₂ component recovery level.

[0034] In accordance with this invention, it is generally advantageous to design the absorbing (rectification) section of the demethanizer to contain multiple theoretical separation stages. However, the benefits of the present invention can be achieved with as few as one theoretical stage, and it is believed that even the equivalent of a fractional theoretical stage may allow achieving these benefits. For instance, all or a part of the pumped condensed liquid (stream **44a**) leaving reflux separator **22** and all or a part of the expanded substantially condensed stream **35b** from expansion valve **14** can be combined (such as in the piping joining the expansion valve to the demethanizer) and if thoroughly intermingled, the vapors and liquids will mix together and separate in accordance with the relative volatilities of the various components of the total combined streams. Such commingling of the two streams shall be considered for the purposes of this invention as constituting an absorbing section.

[0035] FIG. 1 represents the preferred embodiment of the present invention for the processing conditions indicated. FIGS. 3 through 8 depict alternative embodiments of the present invention that may be considered for a particular application. Depending on the quantity of heavier hydrocarbons in the feed gas and the feed gas pressure, the cooled feed stream 31a leaving heat exchanger 10 may not contain any liquid (because it is above its dewpoint, or because it is above its cricondenbar). In such cases, separator 11 shown in FIGS. 1 and 3 through 8 is not required, and the cooled feed stream can be divided into streams 34 and 36, which then can flow to heat exchange (stream 34) and to an appropriate expansion device (stream 36), such as work expansion machine 15.

[0036] As described earlier, the distillation vapor stream 42 is partially condensed and the resulting condensate used to absorb valuable C₃ components and heavier components from the vapors rising through absorbing section 19a of demethanizer 19 (FIGS. 1 and 4 through 8) or absorber column 18 (FIG. 3). However, the present invention is not limited to this embodiment. It may be advantageous, for instance, to treat only a portion of these vapors in this manner, or to use only a portion of the condensate as an absorbent, in cases where other design considerations indicate portions of the vapors or the condensate should bypass absorbing section 19a of demethanizer 19. Some circumstances may favor total condensation, rather than partial condensation, of distillation stream 42 in heat exchanger 13. Other circumstances may favor that distillation stream 42 be a total vapor side draw from fractionation column 19 rather than a partial vapor side draw.

[0037] In the practice of the present invention, there will necessarily be a slight pressure difference between demethanizer 19 and reflux separator 22 which must be taken into account.

If the distillation vapor stream **42** passes through heat exchanger **13** and into reflux separator **22** without any boost in pressure, the reflux separator shall necessarily assume an operating pressure slightly below the operating pressure of demethanizer **19**. In this case, the liquid stream withdrawn from the reflux separator can be pumped to its feed position(s) in the demethanizer. An alternative is to provide a booster blower for distillation vapor stream **42** to raise the operating pressure in heat exchanger **13** and reflux separator **22** sufficiently so that the liquid stream **44** can be supplied to demethanizer **19** without pumping.

[0038] The high pressure liquid (stream **33** in FIGS. 1 and 3 through 8) need not be expanded and fed to a mid-column feed point on the distillation column. Instead, all or a portion of it may be combined with the portion of the separator vapor (stream **34**) flowing to heat exchanger **13**. (This is shown by the dashed stream **38** in FIGS. 1 and 3 through 8.) Any remaining portion of the liquid may be expanded through an appropriate expansion device, such as an expansion valve or expansion machine, and fed to a mid-column feed point on the distillation column (stream **39b** in FIGS. 1 and 3 through 8). Stream **39** in FIGS. 1 and 3 through 8 may also be used for inlet gas cooling or other heat exchange service before or after the expansion step prior to flowing to the demethanizer, similar to what is shown by the dashed stream **39a** in FIGS. 1 and 3 through 8.

[0039] In accordance with this invention, the splitting of the vapor feed may be accomplished in several ways. In the processes of FIGS. 1 and 3 through 8, the splitting of vapor occurs following cooling and separation of any liquids which may have been formed. The high pressure gas may be split, however, prior to any cooling of the inlet gas or after the cooling of

the gas and prior to any separation stages. In some embodiments, vapor splitting may be effected in a separator.

[0040] FIG. 3 depicts a fractionation tower constructed in two vessels, absorber column 18 and stripper column 19. In such cases, the overhead vapor (stream 53) from stripper column 19 may be split into two portions. One portion (stream 42) is routed to heat exchanger 13 to generate reflux for absorber column 18 as described earlier. Any remaining portion (stream 54) flows to the lower section of absorber column 18 to be contacted by expanded substantially condensed stream 35b and reflux liquid (stream 45). Pump 26 is used to route the liquids (stream 51) from the bottom of absorber column 18 to the top of stripper column 19 so that the two towers effectively function as one distillation system. The decision whether to construct the fractionation tower as a single vessel (such as demethanizer 19 in FIGS. 1 and 4 through 8) or multiple vessels will depend on a number of factors such as plant size, the distance to fabrication facilities, etc.

[0041] Some circumstances may favor withdrawing all of the cold liquid distillation stream 40 leaving absorbing section 19a in FIGS. 1 and 4 through 8 or absorber column 18 in FIG. 3 for heat exchange, while other circumstances may not favor withdrawing and using stream 40 for heat exchange at all, so stream 40 in FIGS. 1 and 3 through 8 is shown dashed. Although only a portion of the liquid from absorbing section 19a can be used for process heat exchange when operating the present invention to recover a large fraction of the ethane in the feed gas without reducing the ethane recovery in demethanizer 19, more duty can sometimes be obtained from these liquids than with a conventional side reboiler using liquids from stripping section 19b. This is because the liquids in absorbing section 19a of demethanizer 19 are

available at a colder temperature level than those in stripping section 19b. This same feature can be accomplished when fractionation tower 19 is constructed as two vessels, as shown by dashed stream 40 in FIG. 3. When the liquids from absorber column 18 are pumped as in FIG. 3, the liquid (stream 51a) leaving pump 26 can be split into two portions, with one portion (stream 40) used for heat exchange and then routed to a mid-column feed position on stripper column 19 (stream 40a). Any remaining portion (stream 52) becomes the top feed to stripper column 19. As shown by dashed stream 46 in FIGS. 1 and 3 through 8, in such cases it may be advantageous to split the liquid stream from reflux pump 23 (stream 44a) into at least two streams so that a portion (stream 46) can be supplied to the stripping section of fractionation tower 19 (FIGS. 1 and 4 through 8) or to stripper column 19 (FIG. 3) to increase the liquid flow in that part of the distillation system and improve the rectification of stream 42, while the remaining portion (stream 45) is supplied to the top of absorbing section 19a (FIGS. 1 and 4 through 8) or to the top of absorber column 18 (FIG. 3).

[0042] The disposition of the gas stream remaining after recovery of the liquid co-product stream (stream 47 in FIGS. 1 and 3 through 8) before it is supplied to heat exchanger 60 for condensing and subcooling may be accomplished in many ways. In the process of FIG. 1, the stream is heated, compressed to higher pressure using energy derived from one or more work expansion machines, partially cooled in a discharge cooler, then further cooled by cross exchange with the original stream. As shown in FIG. 4, some applications may favor compressing the stream to higher pressure, using supplemental compressor 59 driven by an external power source for example. As shown by the dashed equipment (heat exchanger 24 and discharge cooler 25) in FIG. 1, some circumstances may favor reducing the capital cost of the

facility by reducing or eliminating the pre-cooling of the compressed stream before it enters heat exchanger 60 (at the expense of increasing the cooling load on heat exchanger 60 and increasing the power consumption of refrigerant compressors 64, 66, and 68). In such cases, stream 49a leaving the compressor may flow directly to heat exchanger 24 as shown in FIG. 5, or flow directly to heat exchanger 60 as shown in FIG. 6. If work expansion machines are not used for expansion of any portions of the high pressure feed gas, a compressor driven by an external power source, such as compressor 59 shown in FIG. 7, may be used in lieu of compressor 16. Other circumstances may not justify any compression of the stream at all, so that the stream flows directly to heat exchanger 60 as shown in FIG. 8 and by the dashed equipment (heat exchanger 24, compressor 16, and discharge cooler 25) in FIG. 1. If heat exchanger 24 is not included to heat the stream before the plant fuel gas (stream 48) is withdrawn, a supplemental heater 58 may be needed to warm the fuel gas before it is consumed, using a utility stream or another process stream to supply the necessary heat, as shown in FIGS. 6 through 8. Choices such as these must generally be evaluated for each application, as factors such as gas composition, plant size, desired co-product stream recovery level, and available equipment must all be considered.

[0043] In accordance with the present invention, the cooling of the inlet gas stream and the feed stream to the LNG production section may be accomplished in many ways. In the processes of FIGS. 1 and 3 through 8, inlet gas stream 31 is cooled and condensed by external refrigerant streams and flashed separator liquids. However, the cold process streams could also be used to supply some of the cooling to the high pressure refrigerant (stream 71a). Further, any stream at a temperature colder than the stream(s) being cooled may be utilized. For instance, a

side draw of vapor from fractionation tower 19 in FIGS. 1 and 4 through 8 or absorber column 18 in FIG. 3 could be withdrawn and used for cooling. The use and distribution of tower liquids and/or vapors for process heat exchange, and the particular arrangement of heat exchangers for inlet gas and feed gas cooling, must be evaluated for each particular application, as well as the choice of process streams for specific heat exchange services. The selection of a source of cooling will depend on a number of factors including, but not limited to, feed gas composition and conditions, plant size, heat exchanger size, potential cooling source temperature, etc. One skilled in the art will also recognize that any combination of the above cooling sources or methods of cooling may be employed in combination to achieve the desired feed stream temperature(s).

[0044] Further, the supplemental external refrigeration that is supplied to the inlet gas stream and to the feed stream to the LNG production section may also be accomplished in many different ways. In FIGS. 1 and 3 through 8, boiling single-component refrigerant has been assumed for the high level external refrigeration and vaporizing multi-component refrigerant has been assumed for the low level external refrigeration, with the single-component refrigerant used to pre-cool the multi-component refrigerant stream. Alternatively, both the high level cooling and the low level cooling could be accomplished using single-component refrigerants with successively lower boiling points (i.e., "cascade refrigeration"), or one single-component refrigerant at successively lower evaporation pressures. As another alternative, both the high level cooling and the low level cooling could be accomplished using multi-component refrigerant streams with their respective compositions adjusted to provide the necessary cooling temperatures. The selection of the method for providing external refrigeration will depend on a

number of factors including, but not limited to, feed gas composition and conditions, plant size, compressor driver size, heat exchanger size, ambient heat sink temperature, etc. One skilled in the art will also recognize that any combination of the methods for providing external refrigeration described above may be employed in combination to achieve the desired feed stream temperature(s).

[0045] Subcooling of the condensed liquid stream leaving heat exchanger 60 (stream 49d in FIGS. 1 and 3, stream 49e in FIG. 4, stream 49c in FIG. 5, stream 49b in FIGS. 6 and 7, and stream 49a in FIG. 8) reduces or eliminates the quantity of flash vapor that may be generated during expansion of the stream to the operating pressure of LNG storage tank 62. This generally reduces the specific power consumption for producing the LNG by eliminating the need for flash gas compression. However, some circumstances may favor reducing the capital cost of the facility by reducing the size of heat exchanger 60 and using flash gas compression or other means to dispose of any flash gas that may be generated.

[0046] Although individual stream expansion is depicted in particular expansion devices, alternative expansion means may be employed where appropriate. For example, conditions may warrant work expansion of the substantially condensed feed stream (stream 35a in FIGS. 1 and 3 through 8). Further, isenthalpic flash expansion may be used in lieu of work expansion for the subcooled liquid stream leaving heat exchanger 60 (stream 49d in FIGS. 1 and 3, stream 49e in FIG. 4, stream 49c in FIG. 5, stream 49b in FIGS. 6 and 7, and stream 49a in FIG. 8), but will necessitate either more subcooling in heat exchanger 60 to avoid forming flash vapor in the expansion, or else adding flash vapor compression or other means for disposing of the flash vapor that results. Similarly, isenthalpic flash expansion may be used in lieu of work expansion

for the subcooled high pressure refrigerant stream leaving heat exchanger 60 (stream 71c in FIGS. 1 and 3 through 8), with the resultant increase in the power consumption for compression of the refrigerant.

[0047] It will also be recognized that the relative amount of feed found in each branch of the split vapor feed will depend on several factors, including gas pressure, feed gas composition, the amount of heat which can economically be extracted from the feed, the hydrocarbon components to be recovered in the liquid co-product stream, and the quantity of horsepower available. More feed to the top of the column may increase recovery while decreasing power recovered from the expander thereby increasing the recompression horsepower requirements. Increasing feed lower in the column reduces the horsepower consumption but may also reduce product recovery. The relative locations of the mid-column feeds may vary depending on inlet composition or other factors such as desired recovery levels and amount of liquid formed during inlet gas cooling. Moreover, two or more of the feed streams, or portions thereof, may be combined depending on the relative temperatures and quantities of individual streams, and the combined stream then fed to a mid-column feed position.

[0048] While there have been described what are believed to be preferred embodiments of the invention, those skilled in the art will recognize that other and further modifications may be made thereto, e.g. to adapt the invention to various conditions, types of feed, or other requirements without departing from the spirit of the present invention as defined by the following claims.